

Liquid-Film Heat-Transfer Coefficients
in a
Forced-Circulation Evaporator

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Liquid Film Heat-Transfer Coefficients in a Vertical-Tube Forced-Circulation Evaporator

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Film heat-transfer coefficients for liquids in forced convection through 8-foot, 0.75-inch i. d., vertical copper tubes have been measured directly in a semi-commercial evaporator and correlated within ± 10 per cent by means of the following dimensionless equation:

$$hD/k = 0.0205 (Du\rho/\mu)^{0.8} (C\mu/k)^{0.4}$$

This equation applies with the above accuracy between the following limits: Reynolds' number, 7500 to 250,000; Prandtl's number, 2 to 50; liquor inlet velocity, 7 to 17 feet per second; viscosity, 1 to 20 pounds per hour per foot.

properties of the liquid, and an equation for liquid film coefficients based on the physical properties has been found to have wide application. Consequently this investigation was conducted for the purpose of obtaining a correlation of the liquid film coefficients in evaporators by means of an equation involving the physical properties of the liquid being evaporated.

THE available data on heat transfer through liquid films deal almost entirely with nonboiling liquids. Some information may be found in the literature concerning liquid film coefficients for boiling liquids. The latest contributions in this field were made in 1931 by Jacob and Fritz (5) and in 1932 by Cryder and Gilliland (3). Linden and Montillon (7) also measured the individual film coefficients in an inclined-tube natural-circulation evaporator. Both of these investigations showed that with natural circulation the values of the liquid film coefficients increase with increases in temperature drop.

There is also considerable information available on over-all coefficients in evaporators. Badger and Shepard (1) and Linden and Montillon (7) found that over-all coefficients increased with increased boiling temperatures. Claassen (2) and Kerr (6) studied the effect of the concentration of boiling sugar solutions on the over-all coefficients and found that an increase in the percentage solids caused a decrease in the coefficient. It is probable that the changes in over-all coefficients with changes in boiling point and concentration are due mostly to the effect of these variables on the liquid film coefficient.

The earlier investigations of heat-transfer coefficients in evaporators expressed the changes in the coefficient in terms of the previously mentioned variables—namely, temperature drop, boiling point, and concentration. Expressions involving these variables limit the application of the results obtained to the particular liquid being evaporated. A change in the variables causes a change in the physical

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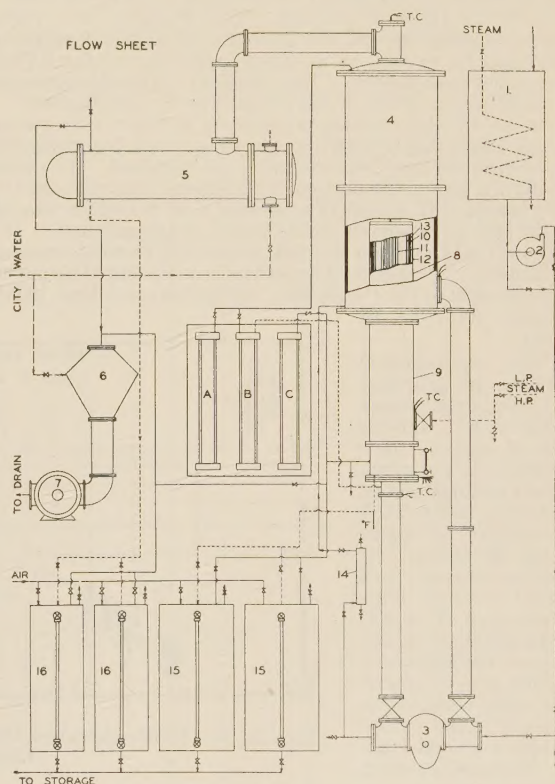


FIGURE 1. EVAPORATOR AND FLOW SHEET

1. 110-gallon feed tank with heating coil
2. Belt-driven centrifugal feed pump
3. Kinney positive pressure circulating pump
4. Main body of evaporator
5. Surface condenser
6. Jet condenser
7. Wet vacuum pump
8. Upper section of heating element
9. Lower section of heating element
10. Asbestos insulation
11. Twelve copper tubes 8 feet long; eleven tubes $\frac{7}{8}$ inch o. d., 16 B. W. G. (Birmingham wire gage) wall; one tube 1 inch o. d., 11 B. W. G. wall
12. Steel shell around upper casting
13. Steam baffle
14. Sampling tube
15. Steam drip tanks
16. Overhead drip tanks
- A. Vacuum manometer with arrangements for automatic control
- B. Pressure drop manometer
- C. Steam pressure manometer

The physical properties that are considered to affect the film heat-transfer coefficient most are the viscosity, thermal conductivity, specific heat, and density of the liquid. In order to obtain a wide range of values of those physical constants, sucrose solutions were used at various concentrations and temperatures. The physical constants were determined from known relationships with the concentration and temperature of the solution.

EXPERIMENTAL PROCEDURE

The experimental unit used was a Swenson semi-commercial forced-circulation evaporator with a heating surface of 18.4 square feet. Figure 1 is a diagram of the evaporator and the flow sheet used for these experiments. Since the upper section of the heating element is surrounded by the boiling liquid, it had to be insulated to prevent the flow of heat through the walls of the casting to the liquid. This was accomplished by building a shell of sheet steel around the heating element and filling the space between the shell and the casting with several layers of asbestos paper. The flow sheet is self-explanatory.

The tube wall thermocouples were installed by the method developed recently by Hebbard and Badger (4). This installation gave consistent and accurate results. Single couples were also placed in the suction line to the pump, in the steam line on the low-pressure side of the control valve, and in the liquor just below the lower tube sheet. A means was devised for measuring the temperature of the liquid in the tubes at intervals from the bottom of the tubes to the top. This consisted of a thermocouple mounted in the end of a 20-foot, 0.25-inch o. d. nickel tube which could be moved up or down inside one of the 0.75-inch tubes in the evaporator. The readings of all thermocouples were made with a Leeds & Northrup type K potentiometer.

It was found while making preliminary test runs that, owing to radiation losses and moisture in the steam from the low-pressure mains, consistent heat balances were not obtained. To decrease the radiation losses, the portion of the steam chest exposed to the air was lagged with a layer of magnesia, and the steam was taken from the high-pressure line at 125 pounds per square inch. It was then found that the steam entering the heating element was always superheated. The degree of superheat was measured by means of a thermocouple in the steam line. All subsequent heat balances checked within 10 per cent.

In the experimental procedure, the principal variables and the ranges covered were: concentration of sugar in solution, 20 to 65 per cent; apparent boiling point of the solution, 160° to 210° F.; velocity of liquid entering the tubes, 7 to 17 feet per second; and apparent over-all temperature drop, 5° to 42° F. The system of changing these variables was such that sets of experiments could be found in which any three of the variables were constant, while the fourth covered the entire range.

The experimental runs were all made in pairs. The normal length of a single run was 15 minutes. The steam drips were read at the beginning, at the end of the first run, and at the end of the second run. The second run in each case was merely a continuation of the first but was calculated separately. The potentiometer readings taken during each run were for the liquor inlet, liquor outlet, steam, tube couples, and nine positions of the traveling liquor couple. It was found possible to read each of these two or three times during each double run. Distilled water was introduced as feed at a rate equal to the rate of evaporation. Thus by maintaining the constant liquor level, the concentration of sugar in solution was kept fairly constant during the run. Samples of the solution were taken at the beginning and the end of

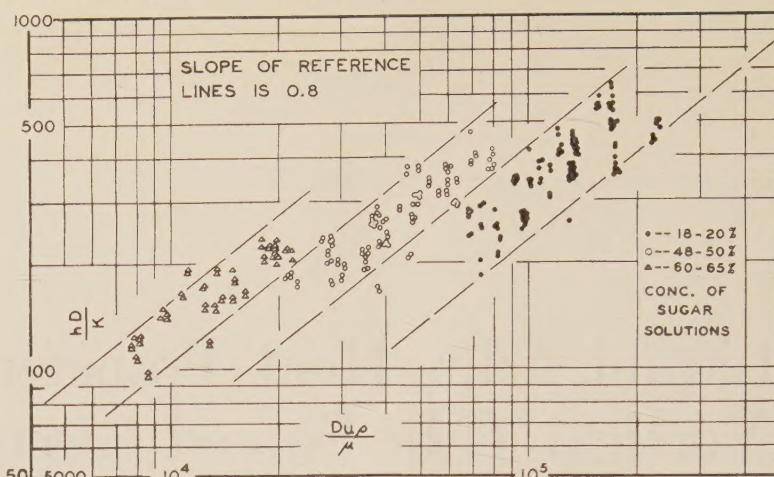


FIGURE 2. DIMENSIONLESS PLOT OF RUNS

each run. These samples were cooled and the percentage sugar determined by means of a Brix hydrometer.

The tube wall temperatures and liquid temperatures were plotted against distance from the bottom of the tube (Figure 5), and the area between the curves was determined by graphical integration. This area, divided by the tube length, gave the true average drop in liquor film temperature.

CORRELATION OF DATA

The data were correlated by the following equation which has been derived by means of dimensional analysis, and is usually known as the Dittus and Boelter equation (8):

$$hD/k = \pi (Dup/\mu)^m (C\mu/k)^n$$

The dimensionless groups were first calculated for each run and plotted as in Figure 2. This plot gave three separate bands for the three series of runs at different concentrations of sugar. The approximate slope of each of these bands is 0.8.

Assuming that the exponent of Reynolds' criterion (Dup/μ) was 0.8, the data were plotted as in Figure 3. This drew all the points into one band which had an approximate slope of 0.4.

Assuming that the exponent of Prandtl's criterion ($C\mu/k$) was 0.4, it was then necessary to recheck the exponent of Reynolds' number. This was done by means of Figure 4. It was found that the best slope for

this band was still 0.8. The values of the exponents are now fixed at $m = 0.8$ and $n = 0.4$. The values of π were calculated for each run, and the average of these values was found to be 0.0205. The final equation is:

$$hD/k = 0.0205 (Dup/\mu)^{0.8} (C\mu/k)^{0.4} \quad (1)$$

Of the 400 runs which were completed, 271 were calculated. Of these 271 runs, 201 were correlated by the above equation to within ± 10 per cent and 117 to within ± 5 per cent.

McAdams (8) recommends the following equation as the

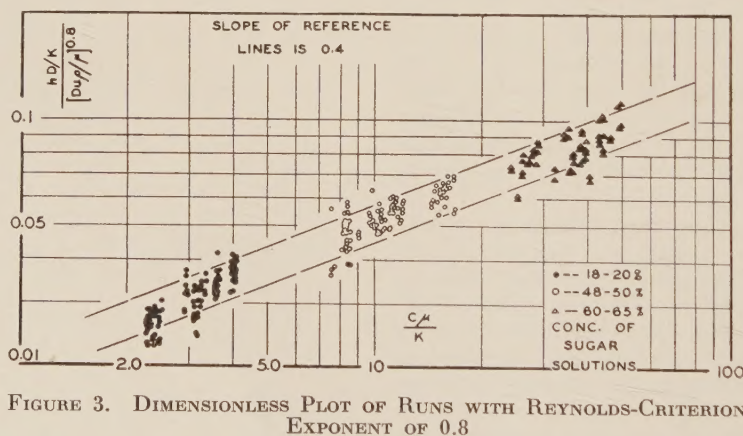


FIGURE 3. DIMENSIONLESS PLOT OF RUNS WITH REYNOLDS-CRITERION EXPONENT OF 0.8

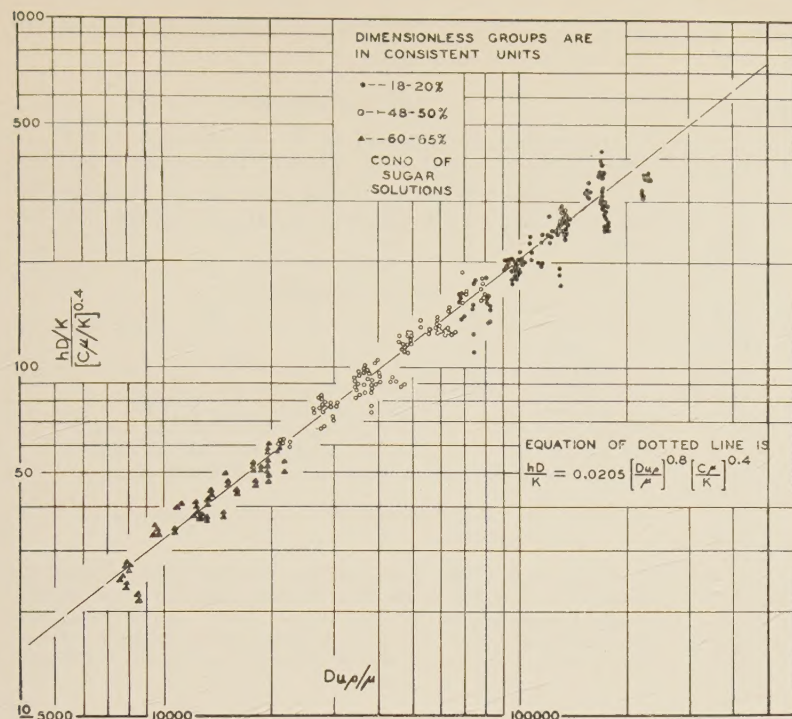


FIGURE 4. PLOT FOR CHECKING REYNOLDS-CRITERION EXPONENT

best general correlation for the heating of nonboiling liquids in turbulent flow in horizontal pipes:

$$hD/k = 0.0225 (Du\rho/\mu)^{0.8} (C\mu/k)^{0.4} \quad (2)$$

The coefficients calculated by means of this equation are slightly higher than those represented by Equation 1. It might seem, therefore, that the insulating effect of the vapors in contact with the tube wall in the boiling section tends to lower the coefficients. It is also logical to assume that the boiling at the top of the tube might increase the velocity and agitation of the liquid to such an extent that the coefficients in this section would be increased.

Figure 5 shows that the liquor temperature curves, in every case, flatten out or reach a maximum at about a foot from the top of the tubes. This might indicate that boiling commences at this point. The vaporization of part of the liquid at this point would result in the sudden withdrawal of heat as latent heat of vaporization. Part of this heat would be removed from the remaining liquid with a resulting lowering of the liquid temperature. This might even result in removing heat from the tube wall at a greater rate than the rate of absorption of heat by the wall from the steam. This would cause a lowering of the wall temperature in this section of the tube. Some of the tube temperature curves also flatten out at the top of the tube.

In this investigation it was found impossible to approximate with any

degree of accuracy the rate of heat transfer to the supposed boiling section. The reason for this was that the length of the nonboiling section is about seven times that of the boiling section, and thus the calculations involved the subtraction of two large quantities which were nearly equal. However, the liquid film coefficients calculated by the equation resulting from this investigation (Equation 1) are so little different from those obtained for nonboiling liquids under the same conditions (Equation 2) that any such effects in the boiling section are of little importance.

The correlation resulting in Equation 1 was not based on theoretical considerations but is merely an empirical means of best representing the experimental results of this investigation. Since Equation 1 is almost entirely empirical, it can be applied only to 8-foot tubes of 0.75 inch inside diameter.

A limitation exists in the application of Equation 1 in the calculation of liquid film heat-transfer coefficients in forced-circulation evaporators. The equation is based on a mean drop in liquid film temperature determined by measuring the change in liquor temperature as it passes through the tube. Only the apparent temperature drop based on the temperature of the liquid in equilibrium with

vapor at the pressure existing in the vapor space of the evaporator can be readily obtained or estimated. Table I shows both the apparent temperature drop and the actual mean

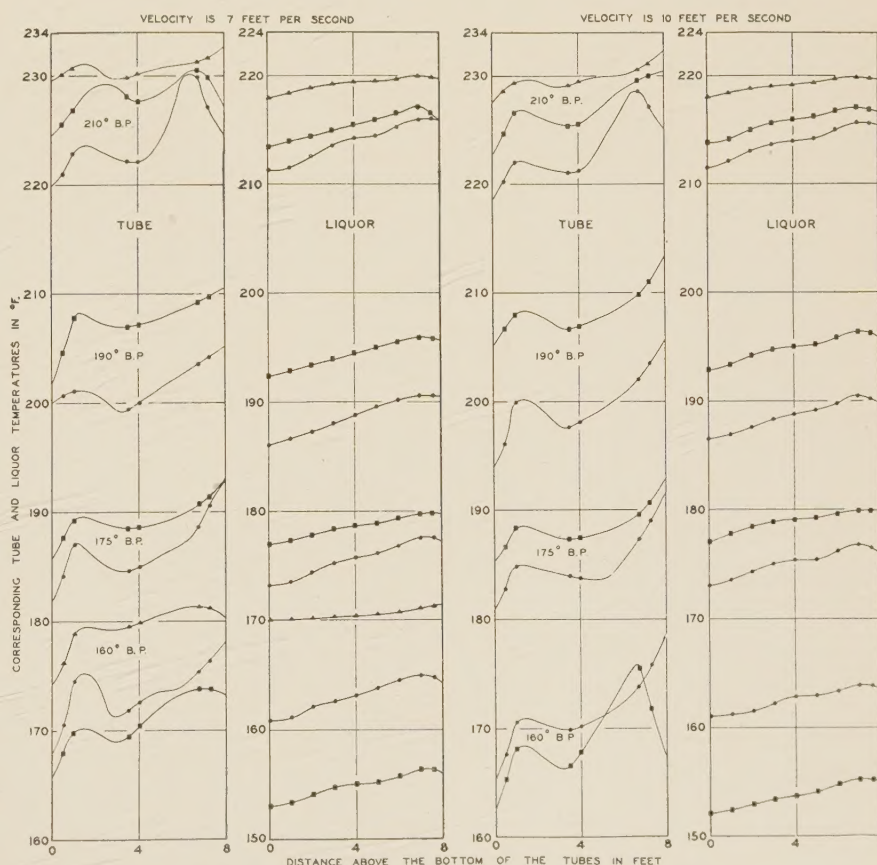


FIGURE 5. TEMPERATURE DISTRIBUTION CURVES

The apparent over-all ΔT for these runs is 20° F. The legend for the concentration of solutions (in per cent sugar) is as follows:

● = 18-20 ■ = 48-50 ▲ = 60-65

temperature drop. No obvious relation seems to exist. The correlation of the liquid film heat-transfer coefficients calcu-

lated on the basis of the apparent temperature drop will be the subject of a later paper.

TABLE I. COMPARISON OF APPARENT AND TRUE TEMPERATURE DROPS

RUN No.	OVER-ALL TEMP.		LIQUID FILM TEMP.	
	Apparent ° F.	True ° F.	Apparent ° F.	True ° F.
AP140	42.1	36.4	24.4	18.7
AP240	42.1	36.6	21.9	16.4
AP340	42.2	38.3	20.6	16.7
AP440	42.3	38.8	18.6	15.1
BP140	42.2	34.6	30.7	23.1
BP240	42.3	35.2	28.9	21.8
BP340	42.3	36.1	25.2	19.0
BP440	42.3	36.9	23.4	18.0
CP140	42.6	32.4	36.0	25.8
CP240	42.6	32.3	33.4	23.1
CP340	42.5	31.2	31.8	20.5
AL140	42.2	36.1	30.4	24.3
AL240	42.6	37.5	22.8	17.7
AL340	42.0	37.5	18.8	14.3
AL442	42.0	38.0	18.2	14.2
BL140	41.5	34.7	31.6	24.8
BL240	42.0	35.4	29.0	22.4
BL340	41.9	39.1	23.7	20.9
BL160	60.4	53.1	45.0	37.7
BL260	62.6	53.5	42.7	33.3
CL144	43.1	33.1	36.7	26.7
CL542	43.2	32.5	34.7	24.0
CL642	43.2	32.1	33.1	22.0

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